COMBUSTION OF LOW GRADE FRACTIONS OF LUBNICA COAL IN FLUIDIZED BED

by

Milica R. MLADENOVIĆ [a*], ***Dragoljub V. DAKIĆ*** [b], ***Stevan DJ. NEMODA*** [a], ***Rastko V. MLADENOVIĆ*** [c], ***Aleksandar M. ERIĆ*** [a], ***Branislav S. REPIĆ*** [a], ***and Mirko S. KOMATINA*** [d]

[a] Laboratory for Thermal Engineering and Energy, Vinča Institute of Nuclear Sciences, University of Belgrade, Belgrade, Serbia
[b] Innovation Center, Faculty of Mechanical Engineering, University of Belgrade, Belgrade, Serbia
[c] Fichtner GmbH & Co. KG, Stuttgart, Germany
[d] Faculty of Mechanical Engineering, University of Belgrade, Belgrade, Serbia

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In this paper a method of examination of fuel suitability for fluidized bed combustion is presented. The research of combustion characteristics of low grade fractions of Lubnica brown coal in the fluidized bed by the aforementioned methodology has been carried out on a laboratory semi-industrial apparatus of 200 kW$_t$. Description of the experimental fluidized bed combustion facility is given, as well as experimental results, with the focus on furnace temperature distribution, in order to determine the location of the zone of intensive combustion. Based on investigation results, which are focused on combustion quality (combustion completion) as well as on satisfying the environmental protection criteria, it can be stated that the investigated coal is suitable for burning in bubbling, as well as in circulating fluidized bed.

Key words*: fluidized bed, coal combustion, low grade fuel, fuel characterisation*

Introduction

The dominant energy potential of Serbia is coal. The majority of Serbian coals is obtained by open-cast mining (about 95%), and is directly used in power plants. Exploitation and mining-geological characteristics of the basin, as well as the need for utilization of coal reserves with heating value below 3500 kJ/kg, prone to slagging and fouling of boiler heat transfer surfaces [1], justifies the use of technology that is less sensitive to oscillations of coal characteristics. A significantly smaller portion of coal is obtained by underground mining, and it is mainly aimed for broad market, and less for power industry. The majority of underground coal mining companies are operating on the verge of profitability for the following reasons:

– undesirable granulation of pit coal – the content of small coal fractions is high, up to 50%, and the coal can hardly be commercially marketed,

* Corresponding author; e-mail: mica@vinca.rs

– often high sulfur and ballast contents, and
– stricter environment protection legislature – adopted legislation with regard to EU directives on permitted emissions from thermal power plants boilers* imposes the necessity to reduce emissions below the level typical for conventional boilers without desulphurization facilities and NO_x reduction measures applied.

These non-commercial fine coal granulations, with adverse chemical composition, often prone to ignition at landfills (which is, apart from being a technical problem, also a major environmental problem), are considered as part of the off-balance coal reserves**. However, they can be a very convenient fuel for local needs. The fact that fluidized bed (FB) boilers can burn fuels with 85% of inert materials, with effective retention of SO_2 by adding limestone into the furnace, and with lower NO_x emissions to meet environmental standards, gives this technology significant advantages compared to other combustion technologies [2, 3]. Therefore, fluidized bed combustion (FBC) technology offers very attractive possibilities for the utilization of poor quality coals, such as Lubnica coal, or similar "difficult" fuels [4]. At the same time, it is one of the ways to increase energy efficiency and environmental acceptability of energy facilities.

Within the scope of the research activities of the Laboratory for Thermal Engineering and Energy of the Vinča Institute of Nuclear Sciences, a method of examination of fuel suitability for FBC has been developed. The main part of the aforementioned methodology is the investigation of fuel combustion on a pilot-scale installation in steady regimes, in order to achieve certain design parameters of a real FB utility or for other purposes [2, 3, 5]. Tests on this experimental facility, besides being cheaper than commercial large-scale boiler tests, are easier to control, and combustion parameters can be modified more easily.

In this paper, investigation of the suitability of Lubnica brown coal for burning in the FB, in terms of gas emissions, combustion stability, and completeness, is presented. This coal is obtained by underground mining and would be used for CHP*** plants near Lubnica coal mines, for heating of the city of Zaječar, as well as for electricity generation. The tests are primarily related to the suitability of this fuel for combustion in a bubbling fluidized bed (BFB). However, on the basis of the analyses from relevant literature listed below, it can be considered that, if this fuel is suitable for combustion in BFB boilers, it will also be suitable for combustion in the circulating fluidized bed (CFB). Namely, differences in combustion conditions between BFB and CFB boilers are a consequence of different CFB hydrodynamics – smaller inert material particle size, higher fluidization velocity, different particle concentration, different mixing, and fuel particle circulation up to the total burn-out. Still, two important parameters for the combustion process are the same: combustion temperature and excess air. There are some more facts that should be kept in mind as well:

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* Guidance on Assessment under the EU Air Quality Directives,
http://ec.erupa.eu/environment/air/pdf/guidanceunderairquality.pdf,
The Law on Air Protection ("Official Gazette of the Republic of Serbia", No 36/2009)
Regulation on Air Quality Requirements and Monitoring Conditions (Official Gazette of the Republic of Serbia, No. 11/2010 and 75/2010),
http://www.ekoplan.gov.rs/aqptwinning/Report/docs/Preliminary%10assessment%20of%20Air%20Quality%20in%20Serbia.pdf

** Out of balance reserves

*** CHP – combined heat and power (also co-generation) integrates the production of usable heat and power (electricity), in one single, highly efficient process

- In both types of boilers, bed temperature is practically the same (800-850 °C), but in CFB boilers it is constant along furnace height, while in BFB boilers a significant difference between bed and freeboard temperature can exist, which depends very much on coal rank (volatile matter content), char reactivity and particle size distribution. However, conditions in the bottom bed of CFB boilers are similar to the conditions in BFB.
- Mixing of fuel particles, gas mixing and inert particle mixing are more intensive in CFB regime, so the convection component of heat transfer is higher in CFB conditions. The height of the CFB boiler furnace is chosen to allow total burn-out of the tiniest fuel particles in one pass. Large particles circulate until they are completely burnt out. Due to these facts, for the same fuel applied, combustion efficiency is higher in CFB boilers than in BFB ones.
- Limestone particle size is smaller, specific surface available for reaction is greater, limestone particles circulate in the furnace permanently, so desulfurization degree is much greater in CFB boilers than in BFB boilers, for the same coal and the same limestone applied.
- Due to the staged combustion, NO_x emission in CFB boilers is smaller than in BFB boilers.

Based on the given short analysis, it can be concluded that if tested coal is suitable for BFB combustion (in terms of environmentally acceptable gas emissions, combustion stability, and completeness), it is also suitable for CFB combustion [6].

Since BFB boilers, due to greater thermal inertia and lower investment costs, have an advantage in the field of small and medium powers up to 50 MW_t and in case emission regulations are not too strict [7], especially in the case of combustion of low quality coals such as Lubnica, a slight advantage might be given to this type of technology.

Materials and methods

The experiment consists of several phases, as follows: (a) preparation of the experimental installation, and basic calculations for operation adjustments, (b) proximate analysis of fuel, (c) test experiments and real operation experiments, (d) ash analysis, and (e) processing and presentation of experimental results.

Before starting the experiments, proximate analysis of coal was performed in order to calculate the adiabatic combustion temperature, as the starting information for adjustments of the experimental installation. The proximate analysis of the fuel is also necessary for setting the mass balance of the operation regime, as well as the heat balance, when required.

Two different temperatures of the FB, 820-830 °C and 850-860 °C, were chosen for performing the experiments, which is the common operating temperature range of industrial FBC facilities. These temperatures are optimal with respect to NO_x concentration, as well as regarding the efficiency of sulfur retention by limestone. Both experiments were conducted in the experimental BFB furnace without a heat exchanger immersed into the bed, and therefore cooled by a considerable amount of excess air which corresponds to the selected adiabatic temperature regimes.

The calculated and real operation parameters are usually different, due to losses through the insulation, losses due to unburnt fuel and other objective reasons. After each working regime, the amount of ash in the separators is measured, which is necessary for setting the mass balance. Ash analysis is done afterwards in order to determine the unburnt combustibles, as well as the particle size distribution of the ash containing the maximum unburnt.

Results of the measurements are processed and presented in diagrams. Measured concentrations of flue gases have been recalculated to the reference oxygen content of 7% [8] in the combustion products, which is required by Serbian legislation. The final activity is drawing the conclusions on the suitability of the given coal for FBC.

Fuel characterization and calculation of the adiabatic combustion temperature

Proximate and ultimate analyses of Lubnica coal are given in tab. 1.

Table 1. Proximate and ultimate analyses of Lubnica coal

	Unit	As receivedd	Analytical	Dry	Dry, ash-free
Moisture	%	30.81	16.40		
Ash		16.16	19.53	23.36	
Sulphur total		1.68	2.03	2.43	
Sulphur in ash		0.48	0.58	0.69	
Sulphur		1.20	1.45	1.73	2.26
Char		40.32	48.72	58.28	45.56
Fixed carbon		24.16	29.19	34.92	45.56
Volatile matter		28.87	34.88	41.72	54.44
Combustible matter		53.02	64.07	76.64	100.00
Fuel heating value					
Higher	kJ/kg	14823	17911	21425	27955
Lower		13625	16916	20685	26991
Ultimate analysis					
Carbon	%	36.58	44.20	52.87	68.99
Hydrogen		2.48	3.00	3.59	4.68
Sulphur		1.20	1.45	1.73	2.26
Nitrogen		0.97	1.17	1.40	1.83
Oxygen		11.79	14.25	17.05	22.24

The calculation of the adiabatic combustion temperature and of the theoretical combustion products volumes (in dry gas) is the starting basis for the determination of the regime parameters of the experimental laboratory-scale FB furnace (calculation of the needed air and fuel flow rates for achieving the desired thermal power, *i. e*. achieving a steady regime at the required temperature). The calculation results for the adopted range of excess air coefficients (λ) are given in tab. 2.

The input data for the calculation are excess air (λ), as well as coal proximate and ultimate analyses (C [%] – carbon, H [%] – hydrogen, O [%] – oxygen, N [%] – nitrogen, S [%] – combustible sulphur, A [%] – ash, W [%] – moisture, LHV [kJkg^{-1}] – lower fuel heating value), and the minimum required amount of air L_0 (kgkg^{-1} of fuel) and V_0 (m^{-3}kg^{-1} of fuel) are calculated.

The output data are:

– TGAS – adiabatic gas temperature [°C]

– GPSS – the amount of dry flue gas [kgkg^{-1} of fuel]
– volume percentages of combustion products [%]: VCO_2, VSO_2, VO_2, VN_2
– mass percentages of combustion products [kgkg^{-1} GPSS]: GCO_2, GSO_2, GO_2, GN_2, GH_2O
– RODG – density of wet flue gas [kgm^{-3}], and
– EGASA – gas enthalpy [kJkg^{-1}].

Laboratory-scale experimental fluidized bed facility

The experimental BFB installation shown in fig. 1 can be used for combustion of solid [9-11] or liquid fuel (with the modified feeding system) [12, 13].

The furnace cross-section area is 300 × 300 mm; the height of the first draft is 2300 mm, while the length of the second draft is 1250 mm. The experiments can be performed with cooling of the FB (when it is required that the excess air during the experiment remains approximately the same to that of real-scale boiler facilities) or without it (in cases when operation with lower excess air is not important, or when design parameters for hot gas generation furnaces are being determined). In the experiments described below the solution without cooling of the FB was used.

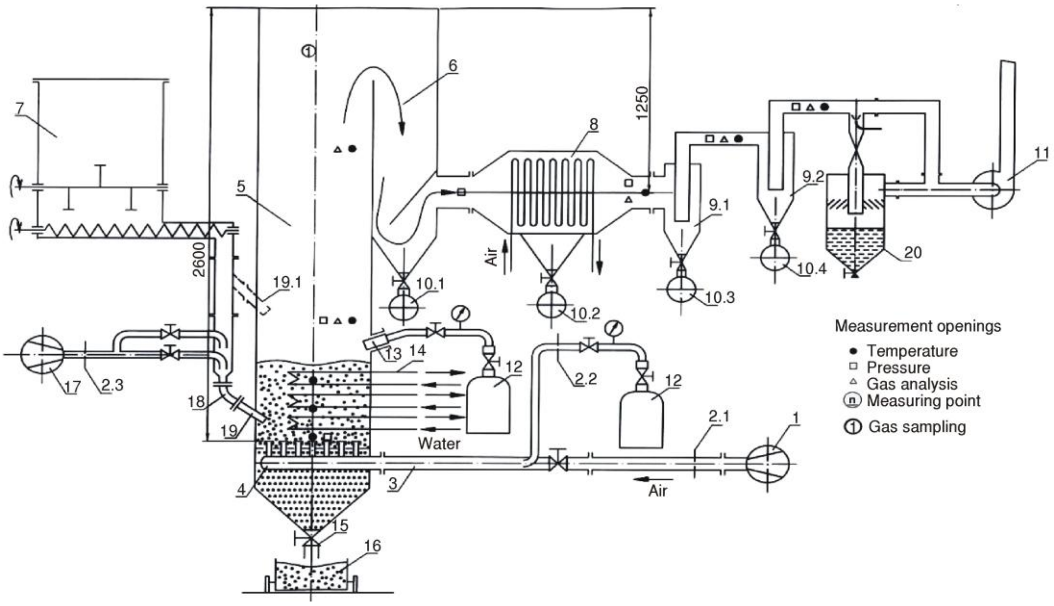

Figure 1. Scheme of the BFB experimental facility with the fuel feeding system
1 – Primary air blower, 2 – (2.1-2.3) Measuring orifices, 3 – Electric heater, 4 – Air chamber with the air distributor, 5 – Fluidized bed furnace (1st draft), 6 – Mechanical device for particle separation (2nd draft), 7 – Fuel feeding, 8 – Flue gases cooler, 9 – (9.1-9.2) Cyclones for particle separation, 10 – (10.1 – 10.4) Vessels for particle disposal, 11 – Flue gas fan, 12 – Propane-butane flask, 13 – Start-up burner, 14 – Three-part heat exchanger, 15 – Furnace material removal valve, 16 – Furnace material collector, 17 – Air blower for pneumatic transport of fuel into the bed, 18 – The line for the visualization of the coal flow, 19 – Coal feeding duct into the bed (19.1 – Coal feeding duct onto the bed), 20 – Scrubber (in operation only when burning toxic materials)

Figure 2. The experimental FB furnace

The fuel is fed onto the bed by a mechanical feeder, and by means of the gravitational force, with maximum granulation (fuel particle size) of 30 mm. Fuel feeding into the bed is performed by pneumatic transport, where the upper size limit for fuel particles is equal to 2 mm.

The photograph shown in fig. 2 demonstrates the lower part of the furnace with the system for pneumatic transport of the coal into the bed, the opening for start-up and overview of the process, as well as the ash particle separator in the second draft (the cylindric vessel in fig. 2).

Equipment used

In order to meet the requirements for regular gas sampling and measurements, the gas analysis system consists of several components (fig. 3). Flue gas sample is taken using the gas sampling probe (1), made of steel resistant to high temperatures. Gas sample passes through the coarse filter (2), which prevents coarser particles to pass further through the line (otherwise, they would cause fouling of gas analyzer cells). The core of the filter consists of inserts (cylindrical plates) made of small pressed bronze spheres. The space between the spheres, originating from the geometrical shape, is large enough to let the gas through, while large ash particles are stopped. The probe is connected with the coarse filter by metal coupling, which provides tightening, in order to avoid the undesired suction of ambient air into the installation.

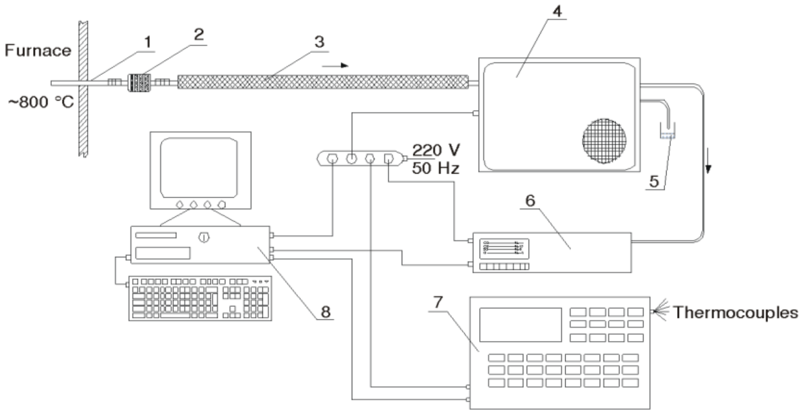

Figure 3. Scheme of the gas analysis and temperature acquisition systems
1 – Gas sampling probe, 2 – Coarse filter, 3 – Heated hose – to prevent the water vapour condensation, 4 – Conditioner, 5 – Condensate removal from the system, 6 – Gas analyzer IMR 3000 P, 7 – Thermocouples and the acquisition system HP3852, 8 – Software for monitoring and registering the measured parameters*

* Hewlett Packard data acquisition and control system

After passing through the coarse filter, the gas enters the heated hose (3), which prevents condensation of water vapour. Presence of water vapour would lead to its reaction with sulfur oxides, creating sulfuric or sulfurous acid, and water would also, due to its own weight, remain in the hose, which would disturb the validity of the analysis. The hose is 5 m long and consists of the central part (through which the gases pass), the heater (covering the hose) and the insulation material. The flexibility of the hose enables its adaptation to different gas sampling conditions. The gas, after particles' separation, passes through the conditioner (4), consisting of the following:

– a filter (for prevention of fine particles to pass further),
– a rapid moisture separator (for removing the moisture in a short period of time, thus preventing the reaction between the moisture and sulfur oxides), and
– a vacuum-pump (which enables gas sample suction and its flowing through the above described system).

During this process, the gas is cooled to the room temperature (around 20 °C).

The conditioner, apart from the flue gas inlet and prepared gas outlet, avails also with a condensate outlet, through which the condensate from the flue gas is removed (5). After conditioning, the flue gas passes through the continuous gas analyser (6) (type IMR 3000 P), which uses electro-chemical sensors for gas analysis, except in the case of CO_2 concentration measurement, which is performed by the NDIR* method.

Temperatures are measured continuously by a system consisting of thermocouples and the HP3852 acquisition system (7). The acquisition system consists of a multi-channel relay multiplexer with thermocouple cold junction compensation, a voltmeter and a data processing system.

Measured data are recorded and continuously monitored online by software, developed in the Laboratory for Thermal Engineering and Energy of the Vinča Institute of Nuclear Sciences.

Results and discusion

Experiments have been done with two different temperatures of the FB:

– Regime I – bed temperature 850-860 °C, and
– Regime II – bed temperature 820-830 °C.

The ground Lubnica coal (size $0 < d \leq 2$ mm) was pneumatically fed into the FB. Total active height of the fixed FB was ≈400 mm, while the coal feeding point was at a 150 mm distance from the bottom of the FB, *i. e.* from the level of the air inlet. In both regimes, temperatures along furnace height have been measured. The thermocouples measuring temperatures t_1, t_2, and t_3 are placed within the FB, where the temperature t_1 is measured practically at the level of the primary air distributor (fig. 4, on the left).

Regime I – bed temperature 850-860 °C

Control of the FB temperature in both regimes of operation has been performed by stopping fuel feeding in short time-intervals, which explains uneven temperature profiles.

* NDIR – A non-dispersive infrared sensor

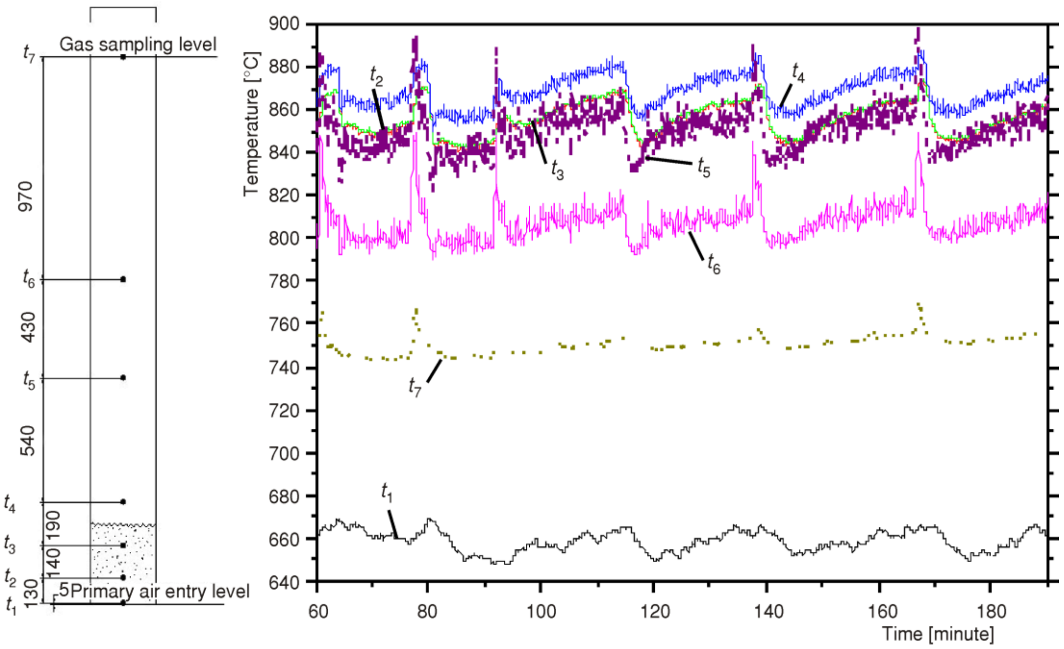

Figure 4. Regime I – temperatures within the furnace (measured values), placement of the thermocouples in the FB furnace – on the left

It is easy to note that temperatures t_2 and t_3 are almost identical during the whole investigation (t_{2avg} = 855.7 °C, t_{3avg} = 856.4 °C), and the temperature t_4 immediately above the fixed bed is only slightly higher than the temperature in the bed itself, which indicates that intense combustion is occurring in the bed or in the "splash" zone.

Namely, average bed temperature of 856 °C corresponds to the minimum fluidization velocity v_{mfI} = 0.323 m/s, calculated according to eq. (1) [14], and to the measured fluidization velocity of v_f = 1.13 m/s (fluidization number N = 3.5):

$$\mathrm{Re}_{\mathrm{mf}} = (24^2 + 0.049\mathrm{Ar})^{0.5} - 24 = \frac{v_{\mathrm{mf}} d_{\mathrm{p}}}{\nu_{\mathrm{f}}} \tag{1}$$

Expanded bed height, *i. e.* expanded splash zone was determined according to the bed expansion ratio eq. (2) [12]:

$$\frac{H}{H_{\mathrm{mf}}} = \frac{1-\varepsilon_{\mathrm{mf}}}{1-\varepsilon} = \frac{\rho_{\mathrm{mf}}}{\rho_{\varepsilon}} \tag{2}$$

$$\varepsilon = \left(\frac{18\mathrm{Re} + 0.36\mathrm{Re}^2}{\mathrm{Ar}}\right)^{0.21} \tag{3}$$

where $\mathrm{Re}_{\mathrm{mf}}$ is the Reynolds number for the particle, at minimum fluidization velocity [–], Re = $= v_f d_p/\nu_f$ – the Reynolds number for the particle [–], Ar = $g d_p^3 \rho_f(\rho_p - \rho f)/\mu_f^2$ – the Archimedes number [–], d_p = 0.8 mm – the particle diameter [m], ν_f – the kinematic viscosity of the fluid (gas) [m^{2}s^{-1}], H – the FB height [m], H_{mf} – the bed height at minimum fluidization [m], ε – the void fraction [–], $\varepsilon_{mf} = 1 - \rho_b/\rho_p = 0.454$ – the void fraction at incipient fluidization [–], ρ_{mf} – the bed density at incipient fluidization [kgm^{-3}], ρ_ε – the fluidized density of FB [kg/m^3], ρ_b = = 1310 – the bulk density of fixed bed [kgm^{-3}], and ρ_p = 2400 – the particle density [kgm^{-3}].

The obtained height of expanded bed from eq. (1) is 510 mm, which indicates that coal combustion occurred in the bed (fig. 10).

High combustion efficiency in the bed has been confirmed also by analyses of the ash collected in the separators and cyclones, as well as by flue gas analyses (figs. 5 and 6). The oscillations of FB temperatures (t_1, t_2, and t_3) were in the range of ±8 °C (~845-860 °C) throughout the tests, while the temperature distur-bances and oscillations are by far more apparent in the space above the bed. By activating the heat exchanger within the bed, temperature control could be performed even more efficiently.

Average oxygen content in the combustion products in Regime I (fig. 5) corresponds to the value of excess air coefficient λ_{avg} = 2.91. This value matches quite well with the value of λ required for achieving the adiabatic combustion temperature of 850 °C (tab. 2). Concentrations of CO, NO_x*, and SO_2 in the combustion products (fig. 6) have been averaged as well and recalculated to the reference value of 7% O_2 in the combustion products, which is stipulated by relevant legislation [8]. A drastic decrease of SO_2 concentration in the final part of the investigation is a consequence of direct limestone feeding onto the bed through the hole at the top of the furnace. This confirms positive effects of limestone feeding into/onto the fluidized bed and the suitability of the FB, considering the possibility of SO_2 emission reduction. These effects could not be quantified completely, due to the lack of limestone analyses data.

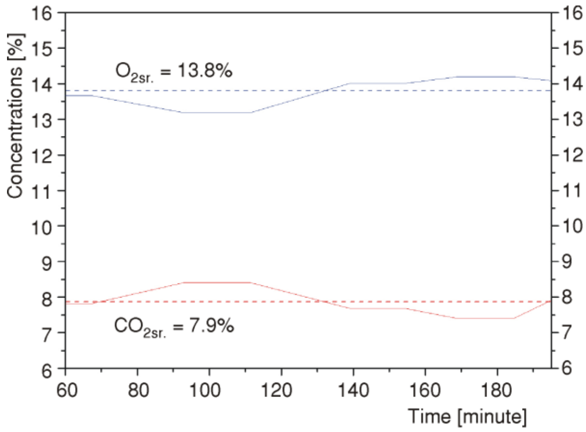

Figure 5. Regime I – concentrations of CO_2 and O_2 (measured and average values)

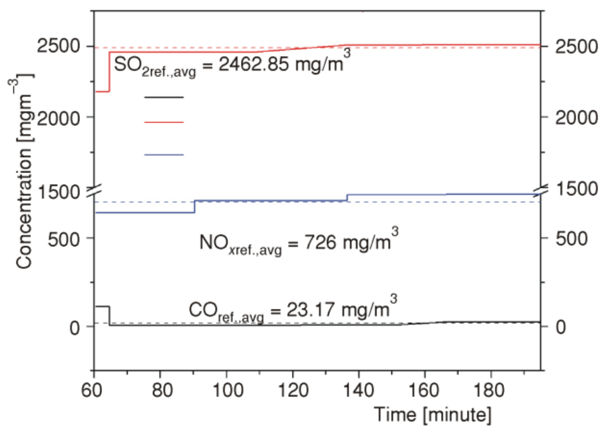

Figure 6. Regime I – concentrations of CO, SO_2, and NO_x at reference O_2 content (O_{2ref} = 7%)

* NO_x is a generic term for the mono-nitrogen oxides NO and NO_2 (nitric oxide and nitrogen dioxide)

Table 2. Calculation of adiabatic combustion temperatures for different excess air coefficients λ

C [%] 36.58	H [%] 2.48	O [%] 11.8	N [%] 0.97	S [%] 1.2	A [%] 16.16	W [%] 30.81		LHV [kJkg^{-1}] 13625	L_0 [kgkg^{-1} of fuel] 4.6394			V_0 [m^{3}kg^{-1} of fuel] 3.548	
λ	TGAS	VCO_2	VSO_2	VO_2	VN_2	GPSS	GCO_2	GSO_2	GO_2	GN_2	GH_2O	RODG	EGAS
1	1866	19.235	0.232	0	80.53	4.95	0.271	0.005	0	0.719	0.117	1.295	13806
1.15	1703	16.684	0.201	2.79	80.33	5.64	0.238	0.004	0.029	0.725	0.104	1.293	13825
1.3	1568	14.73	0.178	4.92	80.17	6.34	0.212	0.004	0.051	0.73	0.093	1.292	13851
1.45	1454	13.186	0.159	6.61	80.05	7.03	0.191	0.003	0.069	0.734	0.085	1.291	13867
1.6	1356	11.935	0.144	7.97	79.95	7.73	0.174	0.003	0.084	0.737	0.078	1.29	13898
1.75	1271	10.9	0.132	9.1	79.87	8.43	0.159	0.003	0.096	0.739	0.073	1.289	13915
1.9	1196	10.031	1.121	10.05	79.8	9.12	0.147	0.003	0.107	0.741	0.068	1.289	13932
2.05	1130	9.29	0.112	10.86	79.74	9.82	0.137	0.02	0.116	0.743	0.064	1.288	13959
2.2	1071	8.651	0.104	11.56	79.69	10.51	0.128	0.002	0.123	0.745	0.06	1.288	13984
2.35	1018	8.094	0.098	12.17	79.64	11.21	0.12	0.002	0.13	0.746	0.057	1.287	13997
2.5	971	7.605	0.092	12.7	79.6	11.91	0.113	0.002	0.136	0.747	0.054	1.287	14021
2.65	927	7.171	0.087	13.18	79.57	12.6	0.107	0.002	0.142	0.748	0.052	1.287	14039
2.8	888	6.785	0.082	13.6	79.54	13.3	0.101	0.002	0.146	0.749	0.05	1.286	14063
2.95	852	6.437	0.078	13.98	79.51	13.99	0.096	0.002	0.151	0.75	0.048	1.286	14078
3.1	819	6.124	0.074	14.32	79.48	14.69	0.091	0.002	0.155	0.751	0.046	1.286	14105
3.25	788	5.84	0.07	14.63	79.46	15.39	0.087	0.002	0.158	0.752	0.044	1.286	14127
3.4	760	5.581	0.067	14.91	79.44	16.08	0.083	0.001	0.161	0.752	0.043	1.285	14139
3.55	734	5.343	0.064	15.17	79.42	16.78	0.08	0.001	0.164	0.753	0.041	1.285	14169
3.7	709	5.126	0.062	15.41	79.4	17.47	0.077	0.001	0.167	0.754	0.04	1.285	14185
3.85	687	4.925	0.059	15.63	79.39	18.17	0.074	0.001	0.17	0.754	0.039	1.285	14204

Regime II – bed temperature 820-830 °C

The temperature history during Regime II is presented in fig. 7.

Since the speed (number of rotations) of the feeding system worm was close to minimum, switching from Regime I to Regime II, *i. e*. lowering the temperature to 820--830 °C, was achieved by more frequent switching off of the feeder, hence oscillations of all temperatures were more obvious than in Regime I. Average values of bed temperatures t_2 and t_3 were almost equal one to another, as it was in Regime I, due to uniform fluidization (t_{2avg} = = 821.4°C, t_{3avg} = 822.3 °C), which can be observed from the diagram in fig. 7.

Average O_2 concentration of 13.9% (fig. 8) corresponds to average excess air coefficient of λ_{avg} = 2.96. This value matches quite well with the value of λ required for achieving the adiabatic combustion temperature (from the calculations – tab. 2). This is considered as a quite good result, taking into account that combustion was controlled by fuel feeding. Namely, both experiments were conducted in the experimental FB furnace without absorption of heat, *i. e*. without a heat exchanger immersed into the bed, which is common in a commercial FB boiler. Therefore, high excess air corresponds approximately to the theoretical combustion temperature of 820-860 °C. High excess air was obtained, for

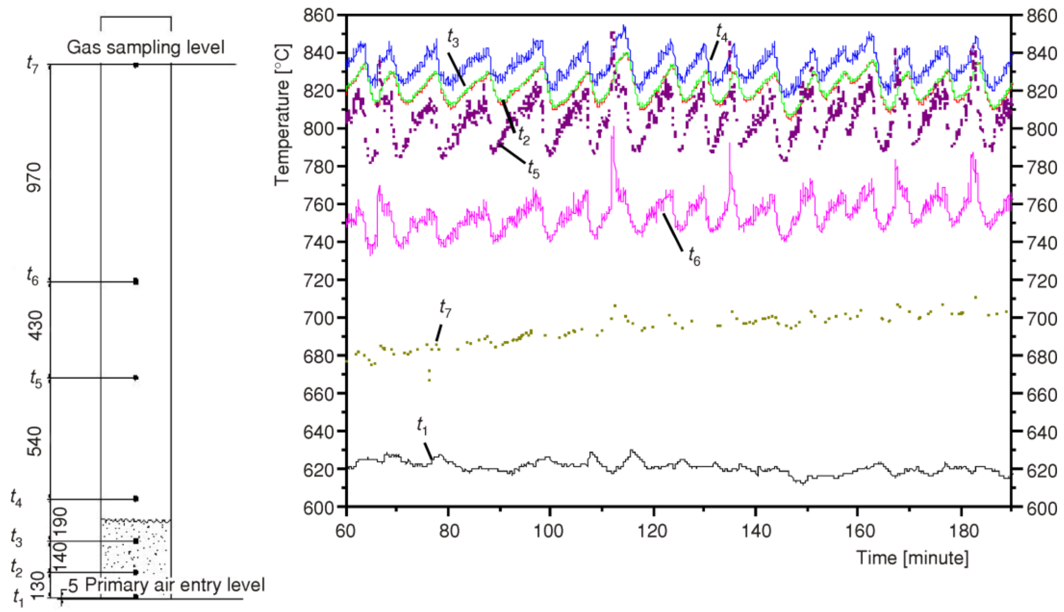

Figure 7. Regime II – temperatures within the furnace (measured values), placement of the thermocouples in the FB furnace – on the left

example, in [15], where bed temperature was controlled in the range of 800-1000 °C, without a heat exchanger in the bed, with the total excess air in the range of 1.1-2.4. In [16], bed temperature was in the range of 754-906 °C and the excess air in the tests was varied between 1.08 and 3.1. Similar experimental results were also obtained in [17]. In case there is a heat sink, *i. e*. existence of heat exchangers immersed into the bed, or cooling water jacket, or flue gas recirculation, the desired bed temperature of 820-860 °C could be achieved/maintained at higher theoretical combustion temperatures and low excess air, which was the case in [18-20].

Average temperature of the expanded bed of 825 °C corresponds to the minimum fluidization velocity v_{mf} = 0.308 m/s and to the measured fluidization velocity of v_f = = 1.148 m/s (N = 3.7). Expanded bed height, determined according to the equation for the bed expansion ratio (2), is H_{exp} = 510 mm, confirming that the combustion zone is located in the bed (fig. 10).

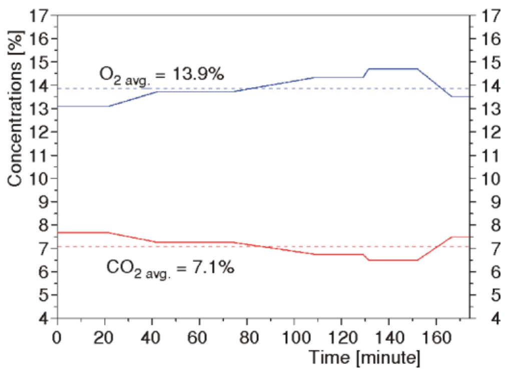

Figure 8. Regime II – CO_2 and O_2 concentrations (measured and average values)

Regime II proved to be more favourable than Regime I, from the point of view of combustion completeness (efficiency), more exactly CO emission, which was cut to half in comparison with Regime I (fig. 9). The emission of NO_x was slightly higher than in Regime I. The concentration of SO_2 in the flue gases was higher in the beginning of the investigation than in Regime I, but it was decreased by introducing limestone into the bed. In

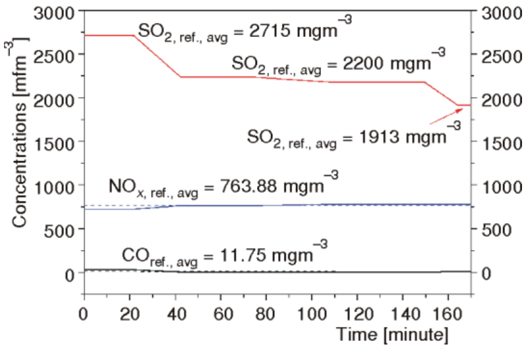

Figure 9. Regime II – CO, SO_2 and NO_x concentrations, at the referent O_2 content (O_{2ref} = 7%)

fig. 9, moments of limestone introduction can be clearly seen, as well as its long-term influence on SO_2 concentration reduction. The data on the amounts and mass flow rates of the fuel and the air, as well as of the ash collected in separators and cyclones, are given in tab. 3. It should be pointed out that, in both working regimes, the mass of the bed material was almost the same before and after the experiment, which showed that there was no ash retention in the bed. On the basis of measured masses of the ash collected in the separators and the cyclones, as well as on the basis of the proximate coal analysis (ash content in the coal, tab. 1), the mass fraction of the fly ash particles emitted into the atmosphere through the chimney was determined.

Table 3. Mass balance of the experimental FB furnace operation in both working regimes

Regime	Fuel mass flow rate	Total mass of the coal used	Collected ash				Primary air flow rate	Secondary air flow rate	Average temperature of the active part of the FB
	kg/h	kg	kg				kg/h	kg/h	°C
			Separator 1	Separator 2	Cyclone 1	Cyclone 2			
I	7.77	31.5	2.32	1.05	0.71	0.29	71.53	18.23	856.0
			∑: 4.37 (86% of the ash) Went out through the chimney: 0.65 (14% of the ash)						
II	8.00	29.74	1.9	1.95	0.5	0.45	67.44	18.23	825
			∑: 4.8 (83% of the ash) Went out through the chimney: 0.74 (17% of the ash)						

Table 4. Fly ash analyses

			Separator 1	Separator 2	Cyclone 1	Cyclone 2
Regime I	Ash	%	98.69	97.22	98.08	96.92
	Combustible		1.31	2.78	1.92	3.08
	Average particle diameter, calculated on total sample mass	μm	157.99	91.66	44.52	42.78
Regime II	Ash	%	98.77	96.98	97.87	96.96
	Combustible		1.23	3.02	2.13	3.04
	Average particle diameter, calculated on total sample mass	μm	159.17	93.88	38.91	38.59

Comparison of average temperature profiles along furnace height, shown in fig. 10, enables the comparison of the regimes. Temperature profiles are very much alike, because of the control method used, as well as due to the design characteristics of the furnace itself.

The investigation of fly ash samples, collected in separators and cyclones, from both working regimes (tab. 4), showed a very low content of combustible matter (mostly below 3%). This proved that most particles almost completely burnt out in the furnace itself.

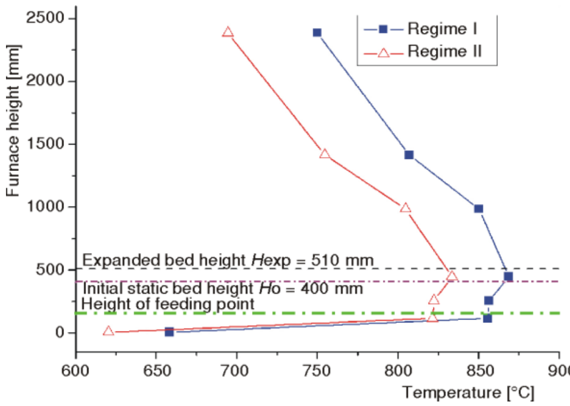

Figure 10. Change of average temperature along furnace height, for both working regimes

Conclusions

The performed investigation of suitability of Lubnica brown coal for FBC was focused on combustion quality, which implies combustion completion, *i. e.* combustion efficiency, and combustion stability as well as satisfying the environmental protection criteria.

As it can be seen from the presented measurement results (diagrams, figs. 6 and 9), CO concentration in the flue gas was considerably lower than the maximum allowable limit stipulated by relevant legislation* (CO $\ll$ 250 mg/m^3). This also implies that losses due to unburnt matter in the gaseous products were negligible. The amount of unburnt matter in the fly ash was also negligible [11].

Based on figs. 4 and 7 can be seen that stable combustion was achieved in both experiments regimes. Temperatures t_2 and t_3 were almost identical during the whole investigation and the temperature t_4 immediately above the fixed bed was only slightly higher than the temperature in the bed itself, which indicates that intense combustion occurred in the bed or in the "splash" zone, in both regimes. Visual observation of the flame through the furnace window has confirmed intensive bed mixing during the experiments, as well good fill of the flame throughout the furnace volume, which indicates that optimal kinetics of the thermal disintegration process of the investigated fuels has been achieved.

High excess air in both regimes is a consequence of the absence of heat exchangers in the FB, which was thus cooled by high air flow. The experiment showed the necessity of cooling the FB to get the excess air required for economical operation of commercial plants.

A comment on combustion quality, with respect to meeting the regulations on environmental protection, is less favourable to some extent. The concentration of SO_2 by far overcomes the maximum permitted limits. During the experiments it was shown that the concentration and emission of sulphur-dioxide could be lowered, relatively easily, by adding

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* According to the *Regulation on the emission limit values of the pollutants in the air (Official Gazette of RS, No. 71/2010)*, the emission limits for boilers with thermal power of the furnace in the range of 1-50 MW with FBC are: CO – 250 mg/m^3, SO_2 – 2000 mg/m^3, and NO_2 – 1000 mg/m^3

limestone. In addition, nitric oxides' concentration in the flue gases was higher than expected. The basic reason for that was high excess air during the experiments, due to the method of controlling fuel feeding and the absence of heat exchangers immersed into the bed. In real conditions, on a large-scale commercial boiler with FB, lower excess air (due to the existence of heat sink) and therefore lower NO_x concentration is expected. Nevertheless, the height of the test facility is normally smaller than the height of the large-scale combustor, which reduces gas residence time. Therefore, it is normal to expect higher NO_x* and CO emission values and potentially more unburnt for a pilot-scale combustor compared to a large-scale plant with similar operational parameters and conditions. Nevertheless, a way for additional decrease of the concentration of nitrogen compounds in the flue gases has to be provided, by adding ammonia or in some other way.

Smaller diameter of a small-scale vessel ensures better radial mixing of the fuel and the fluidizing air, and more uniform distribution of the gaseous compounds inside the combustion chamber [21]. However, this paper put emphasis on the vertical distribution of temperature in the furnace, so the effect of the radial temperature distribution in the narrow combustion chamber was avoided.

Generally, it can be stated that the investigated coal is suitable for burning in bubbling, as well as in circulating FB. It is sufficient to use the advantages of the FB technology, and simple design solutions, in order to meet the law requirements regarding pollutant emissions. Since BFB boilers have an advantage in the field of small and medium powers up to 50 MW_t and since they are less expensive than CFB ones, a slight advantage might be given to this type of technology.

Considering the size of off-balance reserves of lignite in Serbia, as well as large percentage of non-commercial coals from underground mines (about 60% of fine coal fractions), it is possible to build modern, efficient and environmentally friendly boilers with FBC, for production of energy (heat and electricity) in industry and district heating systems, by combusting coals which cannot be burnt in other boiler types [5], or which cannot be burnt efficiently and meet the required environmental standards [8]. At the same time, it is the way to introduce an energy technology, the intense application of which is expected in the 21st century.

Acknowledgments

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* It is a generic term for the mono-nitrogen oxides' NO and NO_2 (nitric oxide and nitrogen dioxide).

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